



Thermodynamic and thermoeconomic analysis of a system with biomass gasification, solid oxide fuel cell (SOFC) and Stirling engine



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ABSTRACT

Thermodynamic and thermoeconomic investigations of a small-scale integrated gasification solid oxide fuel cell (SOFC) and Stirling engine for combined heat and power (CHP) with a net electric capacity of 120 kW_e have been performed. Woodchips are used as gasification feedstock to produce syngas, which is then utilized to feed the anode side of the SOFC stacks. A thermal efficiency of 0.424 LHV (lower heating value) for the plant is found to use 89.4 kg/h of feedstock to produce the above mentioned electricity. Thermoeconomic analysis shows that the production price of electricity is 0.1204 \$/kWh. Furthermore, hot water is considered as a by-product, and the cost of hot water is found to be 0.0214 \$/kWh. When compared to other renewable systems of similar scales, this result shows that if both SOFC and Stirling engine technology enter the commercialization phase, then they can deliver electricity at a cost that is competitive with the corresponding renewable systems of the same size.

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1. Introduction

Due to the ever-increasing demand for more efficient power production and distribution, the main areas of electricity production research and development are efficiency enhancements and pollutant reduction, especially carbon dioxide reduction. There is an increased interest in developing a distributed system of smaller-scale facilities rather than using one large-scale capacity at a specified location, which would allow electricity and heat to be produced and distributed close to the end user, minimizing the costs associated with transportation [1,2].

SOFC stacks will soon enter the commercialization phase, while small Stirling engines are approaching this phase. Therefore, it would be interesting to combine these two technologies into a single system, quantifying the benefits of each system to establish a new technology. Together with an integrated gasification plant that gasifies woodchips in a two-step gasification process, electricity and heat could then be produced in an environmentally friendly way.

SOFCs are one of the most promising types of fuel cells, particularly for energy production. They are expected to produce clean electrical energy at high convention rates, with low noise and low pollution emissions [3].

Due to the high exhaust temperatures of SOFCs caused by the high operating temperature of the cells and the fact that the fuel

utilization in the fuel cell never reaches 100% [4], the unreacted fuel can be combusted in a burner, which in turn produces off-gases with even higher temperature that are perfectly suited for use in a heat engine, such as a Stirling engine, for the production of power and heat for domestic purposes.

The numerous studies on SOFC-based power systems in the open literature have suggested high thermal efficiencies. However, the majority of these studies use gas turbines as the bottoming cycle (see, e.g., Refs. [5–7]). A steam turbine has also been used as a bottoming cycle [8], resulting in high plant efficiency. Only a few studies have been carried out with a Stirling engine as the bottoming cycle when a fuel cell cycle is used as topping cycle (see, e.g., Refs. [1,2,9,10]). At present, using Brayton and Rankine cycles as the bottoming cycles seems to be more applicable because of the maturity of these technologies. As development trends suggest that the SOFC operating temperature will decrease in the future, using a gas turbine as bottoming cycle will be less beneficial.

Integrated gasification SOFC systems have also been studied for some time (see, e.g., Refs. [11–15]). However, integrated biomass gasification SOFC–Stirling CHP (combined heat and power) plants have not been investigated in the open literature, forming the basis of this study.

The present work is an analytical study that includes both thermodynamic and thermoeconomic investigations of systems with integrated gasification of woodchips, where the syngas is used as fuel for a SOFC plant, which also functions as the topping cycle

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for a Stirling engine by contributing heat from the exhausted off-gases, see Fig. 1. The system net capacity is 120 kW_e, which is suitable for decentralized CPH plants. The gasifier model used for the analysis is based on a downscaled version of the Viking two-stage gasifier build at DTU-MEK and now located at DTU Risø. This gasifier produces a clean syngas that can be directly fed into an SOFC (for more information on the gasifier plant, see Refs. [16–18]). The SOFC is based on theoretical equations with empirical coefficients from an experimental arrangement, while the Stirling engine parameters are chosen by fitting a validated feasible engine in terms of construction.

2. Methodology

The thermodynamic results in this study are obtained from the Dynamic Network Analysis (DNA) simulation tool (see, e.g., Ref. [19]). This software is a result of an ongoing development process in the Thermal Energy Section of the Mechanical Department of the Technical University of Denmark, which began with a master's thesis project [20]. Since this time, the program has been continuously developed to be generally applicable, covering unique features and hence supplementing other simulation programs.

The program includes a component library, thermodynamic state models for fluids and standard numerical solvers for differential and algebraic equation systems. The component library models include heat exchangers, burners, turbo machinery, dryers, decanters, energy storages, valves and controllers, among others. The thermodynamic state models for fluids cover most of the basic fluids and such compounds as ash and tar for use in energy system analyses.

DNA is a component-based simulation tool, meaning that the model is formulated by connecting components with nodes and adding operating conditions to build up a system. Next, the physical model is converted into a set of mathematical equations and solved numerically. The equations include mass and energy conservation for all components, and the nodes represent the relationships among the thermodynamic properties of the fluids in the system. The total mass balance and energy balance for the entire system is also included to account for heat loss and heat exchange between different components. In addition, the components include a number of constitutive equations representing their physical properties, e.g., heat transfer coefficients for heat exchangers and isentropic efficiencies for compressors and turbines. The program is written in FORTRAN, and users may also contribute additional components and thermodynamic state models to the libraries.

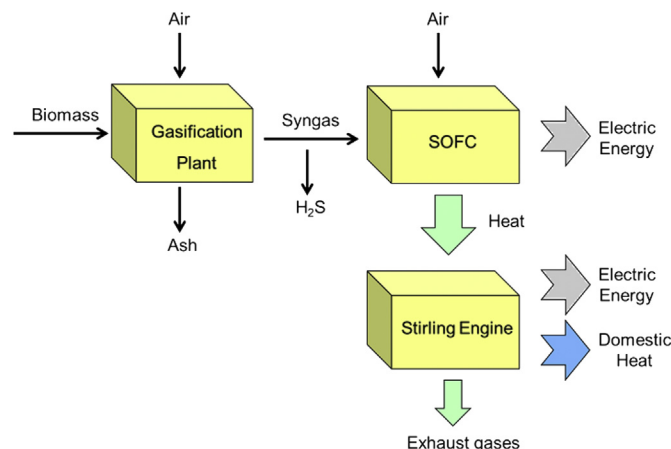


Fig. 1. Block scheme of the plant.

For the thermoeconomy, the general theory is combined with the governing component cost equations to create a system of linear equations, which is then solved in Engineering Equation Solver (EES).

2.1. Modeling of the SOFC stacks

The SOFC model developed in this investigation is based on the planar type developed by DTU Risø and Topsøe Fuel Cell (topsoefuelcell.com). The model was validated against experimental data in the range of 650–800 °C (the range of operating temperatures), as described in Ref. [21]. For clarity, this model is described briefly as follows. The model is assumed to be zero-dimensional, thus enabling the calculation of complicated energy systems. In such modeling, one must distinguish between electrochemical modeling, the calculation of cell irreversibility (cell voltage efficiency) and the calculation of the species compositions at the outlet. For electrochemical modeling, the operational voltage (E_{cell}) is found to be

$$E_{\text{cell}} = E_{\text{Nernst}} - \Delta E_{\text{act}} - \Delta E_{\text{ohm}} - \Delta E_{\text{conc}}, \quad (1)$$

where E_{Nernst} , ΔE_{act} , ΔE_{ohm} and ΔE_{conc} are the Nernst ideal reversible voltage, activation polarization, ohmic polarization and concentration polarization, respectively. Assuming that only hydrogen is electrochemically converted, then the Nernst equation can be written as

$$E_{\text{Nernst}} = \frac{-\Delta g_{\text{f}}^0}{n_e F} + \frac{RT}{n_e F} \ln \left(\frac{p_{\text{H}_2, \text{tot}} \sqrt{p_{\text{O}_2}}}{p_{\text{H}_2\text{O}}} \right), \quad (2)$$

$$p_{\text{H}_2, \text{tot}} = p_{\text{H}_2} + p_{\text{CO}} + 4p_{\text{CH}_4}, \quad (3)$$

where Δg_{f}^0 is the Gibbs free energy (for H_2 reaction) at standard pressure. Because the electrochemical oxidation rate of H_2 is much higher than that of CO , then it can be assumed that hydrogen is the only species that can convert electrochemically (see Refs. [22,23]). In the above equations, p_{H_2} and $p_{\text{H}_2\text{O}}$ are the partial pressures for H_2 and H_2O , respectively.

The activation polarization can be evaluated from the Butler–Volmer equation, which is isolated from other polarizations to determine the charge transfer coefficients and exchange current density from the experiment using the curve fitting technique (see, e.g., Refs. [24,25]).

$$\begin{aligned} \Delta E_{\text{act}} &= \Delta E_{\text{act},c} + \Delta E_{\text{act},a} \\ &= \frac{2RT}{F} \left[\sinh^{-1} \left(\frac{i + i_n}{2i_{0,a}} \right) + \sinh^{-1} \left(\frac{i + i_n}{2i_{0,c}} \right) \right]. \end{aligned} \quad (4)$$

The ohmic polarization depends on the electrical conductivity of the electrodes as well as the ionic conductivity of the electrolyte (see, e.g., Refs. [26,27]). This polarization is also calibrated against experimental data for a cell with anode, electrolyte and cathode thicknesses of 600 μm , 50 μm and 10 μm , respectively.

The concentration polarization is dominant at high current densities for anode-supported SOFCs, wherein insufficient amounts of reactants are transported to the electrodes and the voltage is then reduced significantly. Again, the concentration polarization is calibrated against experimental data by introducing the anode limiting current (see, e.g., Refs. [28,29]), in which the anode porosity and tortuosity are also included, among other parameters.

The fuel composition at the anode outlet is calculated using the Gibbs minimization method, as described in Ref. [30]. Equilibrium at the anode outlet temperature and pressure is assumed for the following species: H_2 , CO , CO_2 , H_2O , CH_4 and N_2 . Thus, the Gibbs

minimization method calculates the compositions of these species at the outlet by minimizing their Gibbs energies. The equilibrium assumption is fair because the methane content in this study is very low. The methane content is mainly depended on the kinetic parameters rather than the chemical equilibrium and its reaction rate is fast [31].

To calculate the voltage efficiency of the SOFC cells, the power production from the SOFC (P_{SOFC}) depends on the amount of chemical energy fed to the anode, the reversible efficiency (η_{rev}), the voltage efficiency (η_v) and the fuel utilization factor (U_F). It is defined in mathematical form as

$$P_{\text{SOFC}} = (\text{LHV}_{\text{H}_2} \dot{n}_{\text{H}_2, \text{in}} + \text{LHV}_{\text{CO}} \dot{n}_{\text{CO}, \text{in}} + \text{LHV}_{\text{CH}_4} \dot{n}_{\text{CH}_4, \text{in}}) \eta_{\text{rev}} \eta_v U_F, \quad (5)$$

where U_F is a set value and η_v is defined as

$$\eta_v = \frac{\Delta E_{\text{cell}}}{E_{\text{Nernst}}}. \quad (6)$$

The reversible efficiency is the maximum possible efficiency, defined as the relationship between the maximum electrical energy available (change in Gibbs free energy) and the fuel's lower heating value (LHV) as follows (see, e.g., Ref. [32]):

$$\eta_{\text{rev}} = \frac{(\Delta \bar{g}_f)_{\text{fuel}}}{\text{LHV}_{\text{fuel}}}, \quad (7)$$

$$\begin{aligned} (\Delta \bar{g}_f)_{\text{fuel}} = & \left[(\bar{g}_f)_{\text{H}_2\text{O}} - (\bar{g}_f)_{\text{H}_2} - \frac{1}{2} (\bar{g}_f)_{\text{O}_2} \right] y_{\text{H}_2, \text{in}} \\ & + \left[(\bar{g}_f)_{\text{CO}_2} - (\bar{g}_f)_{\text{CO}} - \frac{1}{2} (\bar{g}_f)_{\text{O}_2} \right] y_{\text{CO}, \text{in}} \\ & + \left[(\bar{g}_f)_{\text{CO}_2} + 2(\bar{g}_f)_{\text{H}_2\text{O}} - (\bar{g}_f)_{\text{CH}_4} - 2(\bar{g}_f)_{\text{O}_2} \right] y_{\text{CH}_4, \text{in}}. \end{aligned} \quad (8)$$

The partial pressures are assumed to be the average of the inlet and outlet temperatures,

$$\begin{aligned} \bar{p}_j &= \left(\frac{y_{j, \text{out}} + y_{j, \text{in}}}{2} \right) \bar{p} \quad j = \{\text{H}_2, \text{CO}, \text{CH}_4, \text{CO}_2, \text{H}_2\text{O}, \text{N}_2\}, \\ \bar{p}_{\text{O}_2} &= \left(\frac{y_{\text{O}_2, \text{out}} + y_{\text{O}_2, \text{in}}}{2} \right) \bar{p}. \end{aligned} \quad (9)$$

In addition, equations for the conservation of mass (with molar flows), conservation of energy and conservation of momentum are also included in the model. The reliability of the models presented here is justified in Ref. [2] with different fuels such natural gas, ethanol, and methanol.

2.2. Modeling of the Stirling engine

The Stirling engine is noted for its quiet operation and the ease with which it can make use of almost any heat source. Stirling engines are referred to as external combustion heat engines and operate based on a regenerative closed power cycle using helium, nitrogen, air or hydrogen as the working fluid. An ideal regenerative Stirling cycle consists of four processes in one cycle. First, the working fluid absorbs the heat from a high-temperature reservoir and undergoes isothermal expansion (process $1 \rightarrow 2$). Second, the hot working fluid flows through a regenerator, and the regenerator absorbs heat from the hot working fluid. Thus, the temperature of the working fluid decreases in an isochoric process ($2 \rightarrow 3$). Third, the working fluid rejects heat to a low-temperature reservoir and undergoes isothermal compression ($3 \rightarrow 4$). Lastly, the cold

working fluid flows back through the regenerator, and the regenerator rejects heat to the working fluid. The temperature of the working fluid increases in the second isochoric process ($4 \rightarrow 1$).

The model of the Stirling engine provides a rather conservative engine in terms of efficiency, among other properties. It is mainly used to show the applicability of the technology as a functioning bottoming cycle for the usage of a hot exit flow, which in this case is generated by combustion after achieving a rather high temperature in the SOFC topping cycle. The model parameter inputs are selected such that construction is feasible, e.g., infinite surface areas of heat exchangers are not allowed. The heat source used in the analysis is the combustion product gases from the catalytic gas burner, while water used for domestic purposes is used as the sink.

In this study, a pseudo-Stirling cycle, which provides better agreement with engine performance data, is adopted [33]. The power output for the Stirling engine is modeled as

$$P_{\text{Stirling}} = \eta_{\text{pcy}} (Q_{\text{high}} - Q_{\text{loss}}), \quad (10)$$

where Q_{loss} is defined as

$$Q_{\text{loss}} = Q_{\text{high}} (1 - \eta_{\text{mec, stirl}}), \quad (11)$$

where $\eta_{\text{mec, stirl}}$ is the mechanical efficiency of the Stirling engine and Q_{high} is the amount of heat the Stirling engine absorbs from the heat source. The polytropic efficiency η_{pcy} is defined as [33]

$$\eta_{\text{pcy}} = \left[\frac{(1 - \text{RV}^{1-\gamma}) - \zeta(\text{RV}^{\gamma-1} - 1)}{(1 - \text{RV}^{1-\gamma}) + (1 - \zeta)(1 - \varepsilon_{\text{stirl}})} \right], \quad (12)$$

where RV and $\varepsilon_{\text{stirl}}$ are the reversibility factor for the Stirling engine and the effectiveness of the internal heat exchangers in the engine, respectively. The constant γ is equal to 1.667, and ζ is defined as the temperature of the cooler gas over the heater gas,

$$\zeta = \left[\frac{T_{\text{cooler, gas}} + 273.15}{T_{\text{heater, gas}} + 273.15} \right], \quad (13)$$

where $T_{\text{heater, gas}}$ is

$$T_{\text{heater, gas}} = T_{\text{heater, wall}} - \Delta T_{\text{high}}, \quad (14)$$

$$T_{\text{cooler, wall}} = T_{\text{water, inlet}} + 0.66667(\Delta T_{\text{water}}), \quad (15)$$

$$T_{\text{cooler, gas}} = T_{\text{cooler, wall}} + \Delta T_{\text{low}}. \quad (16)$$

2.3. Modeling of the gasifier

The Viking two-stage gasifier is modeled and used for the analysis in this investigation. The Viking gasifier plant is a 75 kW_{th} biomass gasifier using an autothermal (air-blown) fixed bed gasifier. This gasifier is built and developed by the Biomass Gasification Group at the Technical University of Denmark [16]. Wood pellets are used as feedstock. The pellets are first dried to remove surface moisture and pyrolyzed in the first reactor, after which the pyrolytic gases (600 °C) are fed into an autothermal downdraft gasification reactor. The produced exhaust gases after the gasifier are used to heat the reactor for the drying and pyrolysis processes (see Fig. 3). Partial oxidation creates a high-temperature tar-cracking zone in between the pyrolysis and the char gasification step (zone) by injecting preheated air. Char is gasified in the fixed bed, and H₂O and CO₂ are the gasifying agents in the char gasification reactions. The gasifier operates at approximately atmospheric pressure.

The gasifier is modeled by a simple Gibbs reactor, where the Gibbs free energy is at a minimum when chemical equilibrium is achieved. This characteristic is used to calculate the gas composition at a specific temperature and pressure without taking the reaction paths into account [30] and will be briefly explained below. The Gibbs free energy of a gas (assumed to be a mixture of k perfect gases) is

$$\dot{G} = \sum_{i=1}^k \dot{n}_i [g_i^0 + RT \ln(n_i p)], \quad (17)$$

where g_i^0 , R and T are the specific Gibbs free energy, universal gas constant and gas temperature, respectively. Furthermore, each atomic element in the inlet gas is in balance with the outlet gas composition, which shows that the flow of each atom must be conserved. For N elements, this balance is expressed as

$$\sum_{i=1}^k \dot{n}_{i,\text{in}} \mathbf{A}_{ij} = \sum_{m=1}^w \dot{n}_{m,\text{out}} \mathbf{A}_{mj} \quad \text{for } j = 1, N, \quad (18)$$

where the N elements correspond to H_2 , O_2 , N_2 , CO , NO , CO_2 , steam, NH_3 , H_2S , SO_2 , CH_4 , C , NO_2 , HCN (hydrogen cyanide), COS (carbonyl sulfide), Ar and ash (SiO_2) in the gasifying process. $\mathbf{A}_{mj}$ is the number of atoms of element j (H , C , O or N) in each molecule of the entering compound i (H_2 , CH_4 , CO , CO_2 , H_2O , O_2 , N_2 or Ar), while $\mathbf{A}_{ij}$ is the number of atoms of element j in each molecule of the exiting compound m (H_2 , O_2 , N_2 , CO , NO , CO_2 , steam, NH_3 , H_2S , SO_2 , CH_4 , C , NO_2 , HCN (hydrogen cyanide), COS , Ar or ash). The minimization of the Gibbs free energy is formulated by introducing a Lagrange multiplier, μ , for each of the N constraints obtained. The expression can be simplified as

$$\phi = \dot{G}_{\text{tot,out}} + \sum_{j=1}^N \mu_j \left(\sum_{i=1}^k \dot{n}_{i,\text{out}} \mathbf{A}_{ij} - \sum_{m=1}^w \dot{n}_{m,\text{in}} \mathbf{A}_{mj} \right). \quad (19)$$

Furthermore, by setting the partial derivation of this equation with respect to $\dot{n}_{i,\text{out}}$ to zero, the function ϕ can be minimized as

$$\begin{aligned} \frac{\partial \phi}{\partial \dot{n}_{i,\text{out}}} &= \frac{\partial \dot{G}_{\text{tot,out}}}{\partial \dot{n}_{i,\text{out}}} + \sum_{j=1}^N \mu_j \mathbf{A}_{ij} = 0 \quad \text{for } i = 1, k, \\ \Rightarrow g_{i,\text{out}}^0 + RT \ln(n_{i,\text{out}} p_{\text{out}}) + \sum_{j=1}^N \mu_j \mathbf{A}_{ij} &= 0 \quad \text{for } i = 1, k. \end{aligned} \quad (20)$$

Thus, a set of k equations is defined for each chemical compound leaving the system.

In the previous study, Rokni [15] compared the computed dry wood gas composition with the measured dry wood gas composition from Viking gasifier and showed that the computed results agreed very well with the measured data.

2.4. Modeling of the other components

The compressor power consumption is modeled based on the definition of the isentropic efficiency $\eta_{\text{isentropic}}$ and mechanical efficiency $\eta_{\text{mechanical}}$ as

$$\eta_{\text{isentropic}} = \left[\frac{W'}{W} = \frac{\Delta h'_0}{\Delta h_0} = \frac{T'_{\text{exit}} - T_{\text{in}}}{T_{\text{exit}} - T_{\text{in}}} \right]_{\text{compressor}}, \quad (21)$$

$$\eta_{\text{mechanical}} = \left[\frac{\Delta h_0}{W} \right]_{\text{compressor}}, \quad (22)$$

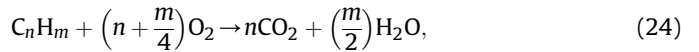
where the mark ' represents the compression process without isentropic losses and W , h and T represent work, enthalpy and temperature.

Heat exchangers are all assumed to counter the flow, and it is assumed that all of the energy is transferred from one side to the other, neglecting heat losses. Both the logarithmic mean temperature difference (LMTD) and the effectiveness-number of transferred units (ε -NTU) methods are used to calculate the flows in the heat exchangers, depending on the type [34].

To model the methanator, which is used to increase the methane content in the fuel, a methanation process is used. This process is mainly expressed by the exothermic reaction of CO and H_2 , which are reformed into CH_4 and steam under high pressure. However, some other minor reactions will also occur.



In the catalytic gas burner, unused fuel is reformed in a highly exothermic process, which is modeled based on the general equation for the burning of hydrocarbons in oxygen,



where n and m denote the amount of the hydrocarbon present in the reaction. Furthermore, for all reforming components, the analysis software uses the Gibbs free energy for the different compounds. Thus, chemical equilibrium is obtained by minimizing the Gibbs free energy for the reactions.

The pump's power consumption is obtained by volumetric and pressure states as

$$W_{\text{pump}} = \left[\frac{\dot{m} v_{\text{in}} (p_{\text{out}} - p_{\text{in}})}{\eta} \right]_{\text{pump}}, \quad (25)$$

where $\dot{m}$, p , v and η are the mass flow, pressure, specific volume (m^3/kg) and efficiency of the pump, respectively.

2.5. Cost model and cost of components

To create the cost model, general theory from Ref. [35] is used to obtain the costs of each separate stream in the energy system and to determine the cost of the produced electricity and domestic hot water, which is considered a by-product. Here by-product means that it is a secondary product which has no influence on financial results and therefore they do not receive allocation of joint costs. The system is modeled using different components, denoted k , with different associated costs. The cost rates of each component are

$$\dot{C}_{P,k} = \dot{C}_{F,k} + \dot{Z}_k^{\text{CI}} + \dot{Z}_k^{\text{OM}}, \quad (26)$$

where $\dot{C}_{P,k}$ is the total cost rate associated with the product of the component, while $\dot{C}_{F,k}$ is the cost rate associated with the fuel (inflow) to the component. $\dot{Z}_k^{\text{CI}}$ and $\dot{Z}_k^{\text{OM}}$ are the cost rates associated with the annual contribution to investment and operation and maintenance, respectively, for component k divided by the annual operation time of the system in hours. Thus, to determine the cost of the entire system, all component cost balances need to be taken into consideration:

$$\dot{C}_P = \dot{C}_F + \dot{Z}_{\text{tot}}^{\text{CI}} + \dot{Z}_{\text{tot}}^{\text{OM}}. \quad (27)$$

The cost rates of the product and fuel are derived from the exergy flow $\dot{E}$ and the specific exergy cost c and are provided below for the exergy transfer rate $\dot{E}_i$, $\dot{E}_e$, power, and heat transfer, expressed in $\$/\text{kWh}$.

$$\dot{C}_i = c_i \dot{E}_i = c_i (\dot{m}_i e_i), \quad \dot{C}_e = c_e \dot{E}_e = c_e (\dot{m}_e e_e), \quad (28)$$

$$\dot{C}_w = c_w \dot{W}, \quad \dot{C}_q = c_q \dot{E}_q. \quad (29)$$

The component cost rates are determined using the total capital investment (TCI) cost, which is found by defining the purchase cost of each component (PEC). The PEC cost equations are obtained from Ref. [36] and are in \$.

$$\begin{aligned} I_{\text{SOFC}} &= I_{\text{stack,SOFC}} + I_{\text{aux,SOFC}}, \\ I_{\text{stack,SOFC}} &= (N_{\text{cell}} \pi D_{\text{cell}} L_{\text{cell}}) (2.96 T_{\text{cell}} - 1907), \\ I_{\text{aux,SOFC}} &= 0.10 I_{\text{SOFC}}, \end{aligned} \quad (30)$$

where cylindrical cells with a diameter of 0.005 m and a length of 0.12 m are assumed. For the SOFC inverter, the cost is

$$I_{\text{SOFC,inverter}} = 10^5 \left(\frac{\dot{W}_{\text{el}}}{500} \right)^{0.7}, \quad (31)$$

($\dot{W}_{\text{el}}$ is in units of W, and the other $\dot{W}$ s, e.g., compressors, are in units of kW).

The cost of the Stirling engine is assumed from Ref. [1] with the assumption that the total TCI is 2200 \$/kW installed. Furthermore, this price represents the present cost and is therefore expected to decrease when Stirling engines reach commercialization levels.

The compressor purchase cost is obtained from

$$I_{\text{compressor}} = 91,562 \left(\frac{\dot{W}_{\text{compressor}}}{445} \right)^{0.67}. \quad (32)$$

The purchase cost of the heat exchangers is obtained from

$$I_{\text{hex}} = 130 \left(\frac{A_{\text{hex}}}{0.093} \right)^{0.78}, \quad (33)$$

and the heat exchanger area is calculate by

$$A_{\text{hex}} = \left[\frac{\dot{m} \Delta h}{k \Delta T_{\text{lm}}} \right], \quad (34)$$

where the values for k are 35 W/m² K for gas–gas heat exchangers and 135 kW/m² K for gas–liquid heat exchangers. ΔT_{lm} is the logarithmic mean temperature difference of the heat exchangers.

For the purchase cost of the fuel reformer, (the gas cleaner and methanator), the following equation is used

$$I_{\text{PR}} = 130 \left(\frac{A_{\text{PR,fin}}}{0.093} \right)^{0.78} + 3240 (V_{\text{PR}})^{0.4} + 2128 V_{\text{PR}}, \quad (35)$$

where the finned area is assumed to be 10% of the volume of the fuel reformer. V_{PR} is obtained from the inlet mass flow and the reaction time of the process. The reaction time is assumed from Ref. [37] to be 1 s for the gas cleaner and 0.5 s for the methanator. The gas burner is assumed to be a catalytic burner, and its cost is therefore assumed to be the same as that for the fuel pre-reformer.

The pump purchase cost is obtained from

$$\begin{aligned} I_{\text{pump}} &= 442 (\dot{W}_{\text{pump}})^{0.71} 1.41 f_{\eta}, \\ f_{\eta} &= 1 + \left(\frac{1-0.8}{1-\eta_{\text{pump}}} \right). \end{aligned} \quad (36)$$

The gasifier purchase cost is assumed from Ref. [38] as follows, considering the small scale of the gasifier:

$$I_{\text{gasifier}} = 1600 (\dot{m}_{\text{drywood}})^{0.67}, \quad (37)$$

where the dry wood mass flow is given in kg/h.

Table 1

Estimates of rates for total investment cost.

Total capital investment	Percentage of PEC
Direct costs	
<i>(a) Onsite costs</i>	
Purchased equipment installation	33%
Piping	35%
Instrumentation and controls	12%
Electrical equipment and materials	13%
<i>(b) Offsite costs</i>	
Civil, structural and architectural work	21%
Service facilities	35%
Indirect costs	
Engineering and supervision	8%
Construction and contractors' profit	15%

The purchase costs of the splitters, mixers and valves are neglected due to the scaling. Thus, the costs related to the other components are considered to be part of the direct costs related to the investment, e.g., piping, instrumentation and controls.

The TCI of PEC for each component k is

$$I_k^{\text{tot}} = I_k \left(1 + \frac{149}{100} \right) \left(1 + \frac{23}{100} \left(1 + \frac{15}{100} \right) \right), \quad (38)$$

where the relationships among the direct and indirect costs related to the investment used are shown in Table 1.

From the TCI of each component, the two cost rates $\dot{Z}_k^{\text{CI}}$ and $\dot{Z}_k^{\text{OM}}$ can be derived. $\dot{Z}_k^{\text{CI}}$ is expressed as

$$\dot{Z}_k^{\text{CI}} = \left[\frac{I_k^{\text{tot}}}{\text{h/year}} \right], \quad (39)$$

where the TCI of component k is amortized in n years by the annuity factor f as

$$\begin{aligned} I_k^{\text{tot}} &= f \times \text{TCI}, \\ f &= \left[\left(\frac{q_i^{(n+\text{CP})} - 1}{(q_i - 1) q_i^{(n+\text{CP})}} \right) - \left(\frac{q_i^{\text{CP}} - 1}{(q_i - 1) q_i^{\text{CP}}} \right) \right]^{-1}, \end{aligned} \quad (40)$$

where q_i is the interest factor obtained from

$$q_i = \left[\left(1 + \frac{\text{int}}{100} \right) \left(1 + \frac{r_i}{100} \right) \right], \quad (41)$$

CP, int and r_i are the construction period, interest rate and rate of inflation, respectively, and are provided in Table 2. For simplification, the lifetime of the whole system is assumed to be 20 years. The construction period is assumed to cover the time from the initial planning until the final completion and commissioning. The interest rate and inflation rates are assumed from Ref. [39], with Denmark used as the area of reference.

Table 2

Economical parameters assumed for the cost rates.

Parameter	Value
Annual operating hours	7500 [h/year]
Interest rate	3%
Inflation rate	2%
Equipment lifetime	20 [years]
Construction period	1 [year]
Maintenance factor	1.1

Furthermore, with M as the maintenance factor, $\dot{Z}_k^{OM}$ can be obtained from

$$\dot{Z}_k^{OM} = \dot{Z}_k^{CI} \times M. \quad (42)$$

This gives the combined contribution to the cost rates associated with the investment of the components.

Since thermoeconomic equations for most components (compressor, turbine, heat exchanger, etc.) are well-known [35], only equations for the gasifier, SOFC and Stirling engine are described here in detail.

A fuel cell integrated with a bottoming cycle can be described by different cost models, varying by *product* and *fuel* allocation [35,36]. In this study the exergy difference between the outgoing used fuel and the inlet reformed gas is considered as *fuel*. Electric power and flue gas are considered as *product*. Cost and exergy balance together with auxiliary equations for SOFC (referring to Fig. 2a) can be found in Table 3.

Similarly, for the Stirling engine (referring to Fig. 2b), exergy and cost balances together with auxiliary equations can also be found in Table 3.

For the gasifier (referring to Fig. 2c), only the syngas is allocated as the product. The fuel is made up of steam-air mixture and dried woodchips. Cost and exergy balances are expressed in Table 3. In order to make the system determined, the auxiliary (steam 99) sets the cost of ash disposal in €/kWh equal to zero.

The average price for woodchips is assumed to be approximately 45 DKK/GJ [40], which is then converted into \$/kWh using the currency rate conversion of 1\$ = 5.7DKK. Transportation costs for the wood are neglected and it is assumed to be obtained locally and therefore have low associated costs. Transportation cost is almost insignificant compared to other cost. In many countries wood are transported through rivers.

By combining all of the component equations, a linear system is created in EES to obtain the costs of all streams of the system. Furthermore, for all components, a single fuel and product are defined for each component so that the cost balances for all components could be found. No heat loss is considered in the thermodynamic analysis; hence, $\dot{E}_q$ (defined in Eq. (29)) is neglected. Thus, the system could be solved by adding a number of auxiliary equations

Table 3
Energy and cost balance for the gasifier, SOFC and Stirling engine.

Gasifier	$\dot{E}_{74} + \dot{E}_2 = \dot{E}_3 + \dot{E}_{99} + \dot{E}_{D, \text{gasifier}}$ $c_{74}\dot{E}_{74} + c_2\dot{E}_2 + \dot{Z}_{\text{gasifier}} = c_3\dot{E}_3 + c_{99}\dot{E}_{99}$ $\dot{E}_{L, \text{gasifier}} = \dot{E}_{99}$ $c_{99} = 0$
SOFC	$\dot{E}_{21} - \dot{E}_{22} = \dot{E}_{28} - \dot{E}_{27} + P_{el} + \dot{E}_{D, \text{SOFC}}$ $c_{21}\dot{E}_{21} - c_{22}\dot{E}_{22} + \dot{Z}_{\text{SOFC}} = c_{28}\dot{E}_{28} - c_{27}\dot{E}_{27} + c_{PeI, \text{SOFC}}P_{el}$ $\dot{E}_{L, \text{SOFC}} = 0$ $c_{21} = c_{22}$ $c_{PeI, \text{SOFC}} = \frac{c_{28}\dot{E}_{28} - c_{27}\dot{E}_{27}}{\dot{E}_{28} - \dot{E}_{27}}$
Stirling engine	$\dot{E}_{40} - \dot{E}_{41} = \dot{E}_{43} - \dot{E}_{42} + P_{el} + \dot{E}_{D, \text{Stirling}}$ $c_{40}\dot{E}_{40} - c_{41}\dot{E}_{41} + \dot{Z}_{\text{Stirling}} = c_{43}\dot{E}_{43} - c_{42}\dot{E}_{42} + c_{PeI, \text{Stirling}}P_{el}$ $\dot{E}_{L, \text{Stirling}} = 0$ $c_{40} = c_{41}$ $c_{PeI, \text{Stirling}} = \frac{c_{43}\dot{E}_{43} - c_{42}\dot{E}_{42}}{\dot{E}_{43} - \dot{E}_{42}}$

based on the assumptions that the fuel flowing through a component can be used in a later stage, which means that the fuel had the same unit cost at the inlet and outlet. If the product of a component is composed by two or more streams, then the unit cost of those streams are assumed to be the same. Thus, for the heat exchangers and dryer, the specific exergy cost of the hot streams is considered to be equal. Furthermore, for compressors and pumps, the specific exergy cost of the inlet ambient air or ambient water is assumed to be zero. For the gas cleaner and gasifier, the disposal costs of sulfur and ash are neglected, meaning that the specific exergy cost is zero. The splitters' outlet streams are assumed to be equal. For SOFC, the assumptions are 1) that the exergy difference of the inlet fuel and the outlet fuel is the fuel consumption of the SOFC and hence had equal specific exergy costs and 2) that the flue gas is considered as waste, and hence the specific exergy cost is zero. In this case, the fuel for Stirling bottoming cycles still had costs related to the streams, and if more fuel is left in the stream (due to the lower fuel utilization factor of the SOFC), then this assumption is reasonable, and in this way, the bottoming cycle is not considered a full regenerative cycle. For the Stirling engine, two auxiliary equations are needed. First, the outlet stream on the hot side is considered as waste and hence zero. Additionally, the water heater utilizing this waste is assumed to recover

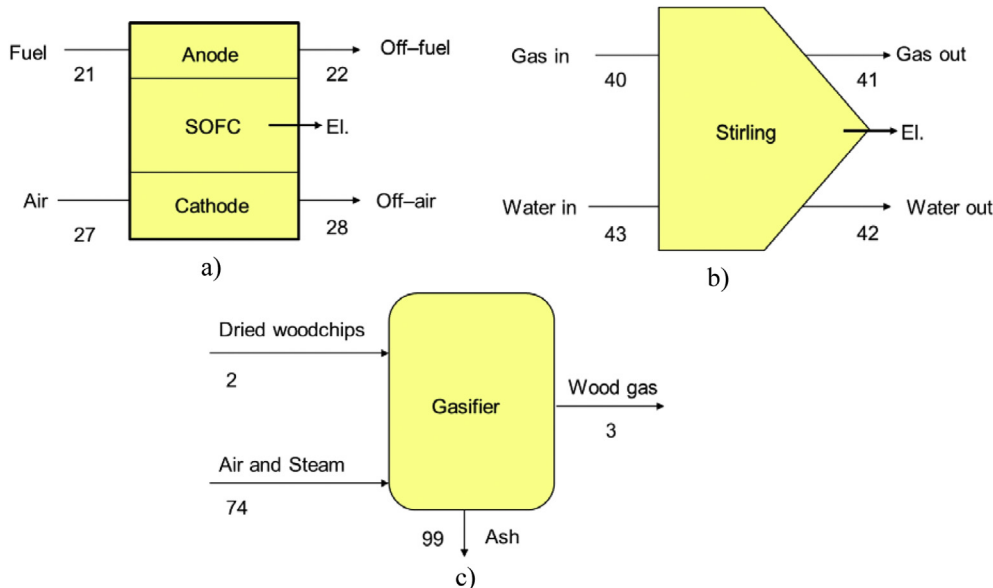


Fig. 2. (a) SOFC, (b) Stirling engine and (c) gasifier streams schemes.

energy from the waste. The second assumption is that the products of the Stirling components are mechanical power from the engine, and the heat absorbed by the heat source thus leads to

$$C_{P,mec} = \left[\frac{c_{cold,in} \dot{E}_{cold,in} - c_{cold,out} \dot{E}_{cold,out}}{\dot{E}_{cold,in} - \dot{E}_{cold,out}} \right]. \quad (43)$$

Electricity is considered the main product of the plant, and heat is considered the secondary product, which will be present as long as electricity is produced. It is thus assumed that the internal power consumption is covered by the production; hence, the internal power cost will be equal to the cost of the produced electricity:

$$Price_{el} = \left[\frac{\dot{E}_{P,SOFC} C_{P,SOFC} + \dot{E}_{P,Stirling} C_{P,Stirling}}{\dot{E}_{P,SOFC} + \dot{E}_{P,Stirling}} \right]. \quad (44)$$

3. Plant configuration

The system investigated herein is a small-scale CHP consisting of an integrated biomass gasification plant with an SOFC system as the topping cycle and a Stirling engine with a water heater as the bottoming cycle (see Fig. 3).

Woodchips are fed into the system to produce syngas. This is performed in a two-step process, wherein the first step is the drying and pyrolysis of the feedstock and the second is fixed bed gasification, where the pyrolyzed feedstock is gasified by steam and air as gasification agents. Although Ref. [18] reported that the produced syngas is clean enough to be fed directly into the SOFC without additional fuel processing, a gas cleaner is introduced for the removal of H_2S present in the syngas and bring its concentration below 1 ppm level. The gas cleaner is assumed to work at a temperature of 250 °C.

For the topping SOFC cycle, ambient air at 15 °C is compressed to the working pressure of the SOFC (normal pressure) before being heated in the cathode air preheater (CP) to the cathode inlet temperature of 600 °C. The cathode preheater uses some of the SOFC off-air for the heating. The off-air is split in two streams: one entering the CP and the other entering the catalytic burner (CB). For the anode side, the cleaned syngas is first pumped to compensate

for the pressure drop along the way. Next, the syngas is sent to the methanator, wherein the molar fraction of CH_4 in the gas is increased from approximately 0.01 to nearly 0.05 at the expense of the molar fractions of H_2 , and CO and steam, while the molar fractions of N_2 and CO_2 also increase. This will not affect the SOFC's electrical production in any particular way; however, because the reformation is highly exothermic, less heat needs to be extracted from the SOFC off-fuel to heat the incoming fuel to the SOFC fuel inlet temperature of 650 °C in the anode preheater (AP). The reformation allows the Stirling engine to receive a larger amount of heat to be used due to the higher temperature of the fuel when entering the CB, causing the combustion processes to occur at a higher temperature. The CB is implemented because not all of the fuel is reacted in the SOFC stacks due to fuel utilization. The entering temperatures are essential requirements for the proper functioning of the SOFC stacks in terms of both initiating the chemical reactions and avoiding cell thermal fractures.

Second, a larger portion of CH_4 in SOFC causes endothermic internal reforming, which leads to the use of less air for cooling to maintain the SOFC operating temperature at 780 °C. Therefore, the workload of the cathode compressor/air blower will decrease.

For the bottoming cycle, a Stirling engine is implemented. The Stirling engine utilizes the combustion products leaving the CB as the heat source. For the heat sink, water is used with an incoming temperature of 20 °C and an outgoing temperature of 60 °C so that it can be used, for example, as domestic hot water for room heating, at a temperature that is sufficient to address bacterial problems (e.g., *Legionella* [41]) and to be useful for domestic purposes. The heat remaining after the Stirling engine is used for domestic water heating. Water is constrained in the same manner as the heat sink, and the combustion products leaves the system into the environment at approximately 95 °C, which is high enough to avoid corrosion problems.

4. Results and discussion

The main operating parameters for the plant are presented in Table 4. The ambient conditions are assumed to be 1 bar and 15 °C. The presented values are the final values used after the optimization.

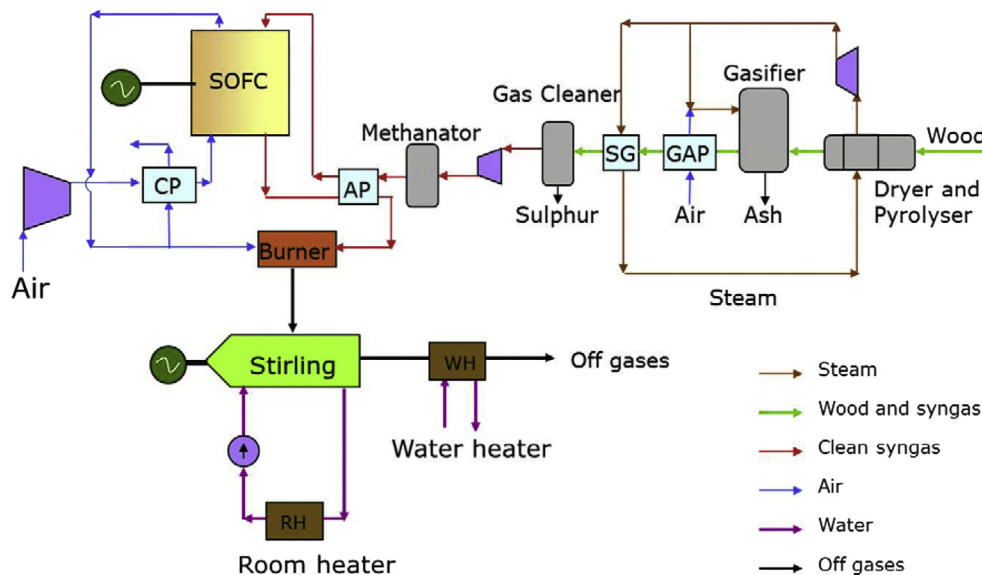


Fig. 3. System layout for the integrated gasification SOFC–Stirling-based CHP plant.

Table 4
System operating input parameters.

Parameter	Value
Woodchip temperature	15 °C
Pyrolysis and dry wood temperature	150 °C
Gasifier temperature	800 °C
Gasifier pressure	1 bar
Gasifier pressure drop	0.005 bar
Gasifier carbon conversion factor	1
Gasifier non-equilibrium methane	0.01
Steam blower isentropic efficiency	0.8
Steam blower mechanical efficiency	0.98
Steam temperature in steam loop	150 °C
Wood gas blower isentropic efficiency	0.7
Wood gas blower mechanical efficiency	0.95
Gas cleaner pressure drop	0.0049 bar
Cathode compressor air intake temperature	25 °C
Compressor isentropic efficiency	0.7
Compressor mechanical efficiency	0.95
SOFC operating temperature	780 °C
Anode inlet temperature	650 °C
Cathode inlet temperature	600 °C
Pressure drop, anode side	0.02 bar
Pressure drop, cathode side	0.055 bar
SOFC fuel utilization rate	0.675
Number of cells per stack	74
Number of stacks	160
Heat exchanger pressure drop	0.01 bar
Pinch temperature CP	20 °C
Burner ratio inlet outlet pressure	0.97
Stirling engine heater wall temperature	600 °C
Stirling engine ΔT_{high}	125 °C
Stirling engine ΔT_{low}	60 °C
Regenerator RV loss factor, Stirling engine	1.44
Heat exchanger efficiency, Stirling engine	0.98
Stirling engine mechanical efficiency	0.8
Stirling engine heat sink inlet temperature	20 °C
Stirling engine heat sink outlet temperature	60 °C
Water pump efficiency	0.95
Water heater inlet water temperature	20 °C
Water heater outlet water temperature	60 °C
Off gas temperature (after water heater)	95 °C

To achieve a 120 kW_e output of electrical energy, the gasifier needs an input of approximately 89.4 kg/h, which leads to a syngas production of approximately 176.4 kg/h. Thus, this amount of biomass must be obtained from agriculture or a cultivation area and provided to the unit. The syngas molar fraction is presented in Table 5.

Because ambient air is used as one of the gasifying agents, a large portion of unusable N₂ is present in the syngas. This issue is addressed by the application of a methanator prior to SOFC without using anode recirculation or external steam supplement. On the other hand, this leads to a larger mass flow, which is also beneficial for the Stirling engine operation. For the SOFC, however, a large amount of N₂ and steam causes a concentration polarization at a

Table 5
Syngas and methanated syngas molar fraction composition.

Compound/molar fraction	Wood	Syngas	Methanated syngas
Hydrogen	0.041	0.253	0.220
Nitrogen	0.001	0.288	0.310
Carbon monoxide		0.172	0.086
Carbon dioxide		0.116	0.186
Water (liquid)	0.332		
Water (gas)		0.158	0.147
H ₂ S ($\times 10^6$)		45.71	—
Methane		0.010	0.049
Argon		0.003	0.004
Oxygen	0.293		
Carbon	0.326		
Sulfur ($\times 10^6$)	133.6		
Ashes	0.006		

rather early stage. The SOFC has a power output of 98.8 kW_e, while the Stirling engine provides 26.9 kW_e of power. The internal power consumption is 5.8 kW, which is mainly due to the cathode air compressor. The high power consumption for the cathode air compressor is due to the cooling effect of this compressor, which is needed to maintain the SOFC temperature at the desired level. The reaction inside the SOFC is highly endothermic and therefore needs a relatively large flow of air to cool the cells.

The thermal efficiency of the topping SOFC cycle is 0.329 LHV, which is somewhat low for a SOFC system (see Table 6). However, the entire plant has a thermal efficiency of 0.424 LHV, which is a decent value for such a small-scale system. The implementation of the bottoming Stirling engine gives a remarkable increase in plant efficiency of 28.9%.

The Stirling engine heat sink produces nearly two thirds of the domestic water heating. The two water heaters together produce 127.33 kW_{th}.

It is also assumed that the plant's internal electrical consumption is covered by its production, meaning that the 120 kW_e production of electricity is the net production,

$$P_{net} = P_{SOFC} + P_{Stirling} - \sum_n P_{el,n,consumed} \quad (45)$$

where P_{SOFC} and $P_{Stirling}$ are the power production from the SOFC and Stirling engine, respectively, and $P_{el,n,consumed}$ is the consumption of power from the n th component. The thermal efficiency of the plant is calculated by the net power production of the SOFC and the Stirling engine compared to the fuel input:

$$\eta_{th,plant} = \left[\frac{P_{net}}{\dot{m}_{fuel} \times LHV_{fuel}} \right] \quad (46)$$

4.1. Thermodynamic investigation

The most important parameters for this investigation are the woodchip mass flow, number of SOFC stacks and SOFC cell utilization factor. The woodchip mass flow indicates the cultivation area to be allocated to provide the mass flow needed by the fuel. The number of SOFC stacks is directly related to the SOFC purchased cost and therefore investment cost, while the utilization factor affects the amount of off-fuel (fuel remaining after the SOFC stacks) that would be available for the bottoming cycle (the Stirling engine in this case). The lower utilization factor means that more fuel will be available for the Stirling engine; thus, the engine will produce more power. That is, the utilization factor affects the cooperation between the two cycles, namely, the SOFC plant and Stirling engine.

Table 6
Plant output for the initial operating parameters.

Parameter	System output
Feedstock consumption	89.4 kg/h
Syngas produced	176.4 kg/h
Power output, SOFC	98.8 kW
Power output, Stirling engine	26.9 kW
Total power consumption	5.8 kW
Thermal efficiency, SOFC cycle	0.329
Thermal efficiency, plant	0.424
Percentage increase when adding Stirling cycle	28.9%
Component	Produced heat
Stirling engine	53.5
Water heater	73.83
Total hot water production	127.33

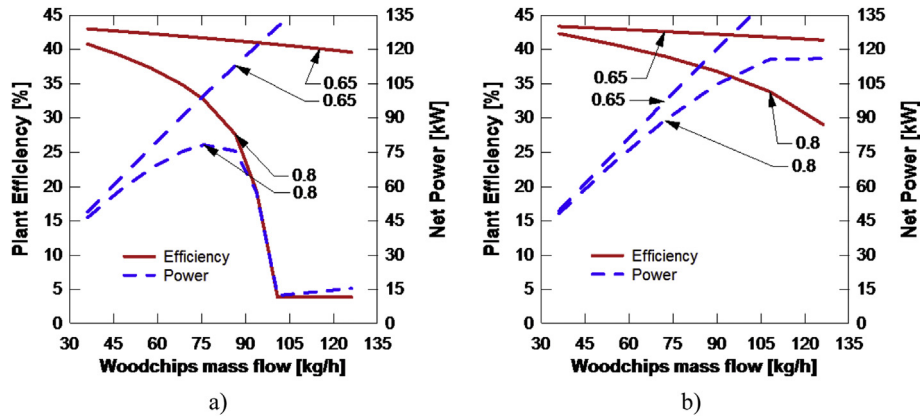


Fig. 4. Plant thermal efficiency and net power production as function of woodchip mass flow with fuel utilization of 0.65 and 0.8: a) 100 stacks and b) 150 stacks.

Following the discussion above, the first parameter to be investigated is the woodchip mass flow. To study the system performance with different woodchip mass flows, the initial analysis is performed on a system using 100 and 150 SOFC stacks with 74 cells per stack. The fuel utilization of the SOFC cells is initially assumed to be 0.8, see Ref. [4]. The final utilization factor will be decided later. The plant efficiency and net power production as functions of woodchip mass flow are shown in Fig. 4 for two different utilization factors as 0.8 and 0.65.

As shown, the thermal efficiency of the system tends to decrease as the mass flow in the system increases, and at a certain level of fuel mass flow, it tends to decrease rapidly. For the case of 100 stacks, the power increases as the woodchip mass flow increases until reaching a maximum, after which it decreases. This behavior is also observed for the case with 150 stacks, but the woodchip mass flow is much higher. For other values of the SOFC utilization factor than 0.8, the results will be slightly different but the overall conclusions remain the same.

Another important parameter is the SOFC utilization factor. To study this parameter, the woodchip mass flow is fixed to produce a net power of 120 kW at a utilization factor of approximately 0.7. Fig. 5 shows the variation of the plant efficiency and net power production when the SOFC utilization factor is changed. The results are shown for two different numbers of stacks (100 and 150) to study the plant performance.

Several interesting points can be concluded from this figure. A utilization factor exists for which the plant efficiency is highest: 0.625 for 100 stacks and 0.65 for 150 stacks. This maximum point is

unique for a given number of stacks and must be calculated using the number of stacks chosen based on an economic analysis. Another issue to be mentioned is that the net power production and plant efficiency decrease sharply after a certain utilization factor. This point is found out to be approximately 0.8 for 100 stacks and 0.83 for 150 stacks. At this point, the concentration losses in the SOFC cells dominate, and as a result, the cell voltage decreases significantly. This point will obviously be shifted to the right when the number of stacks is increased. Increasing the number of stacks decreases the fuel amount for each stack when the total fuel mass flow remains constant.

As mentioned above, the number of stacks is also an important issue to study. For this reason, several calculations are performed, and the results are illustrated in Fig. 6. For comparison purposes, two utilization factors are selected: 0.8, which is a rather high value, and 0.65, which is the optimum value when 150 stacks are used. Again, the fuel mass flow is fixed for these cases. For the case with a utilization factor of 0.8, increasing the stack number from 100 to approximately 120 sharply increases the plant efficiency and power, and further increases in the stack numbers increase plant performance significantly. Eventually, the amount of fuel per stack becomes too large, and the ionic concentration reaches its limit. Increasing the stack number distributes the fixed fuel to more stacks and reduces the concentration to below its limit.

For the case with a utilization factor of 0.65, increasing the stack numbers slightly increases plant efficiency and power. The problem of ionic concentration does not exist for such this lower utilization factor. Another conclusion from these results is that the plant

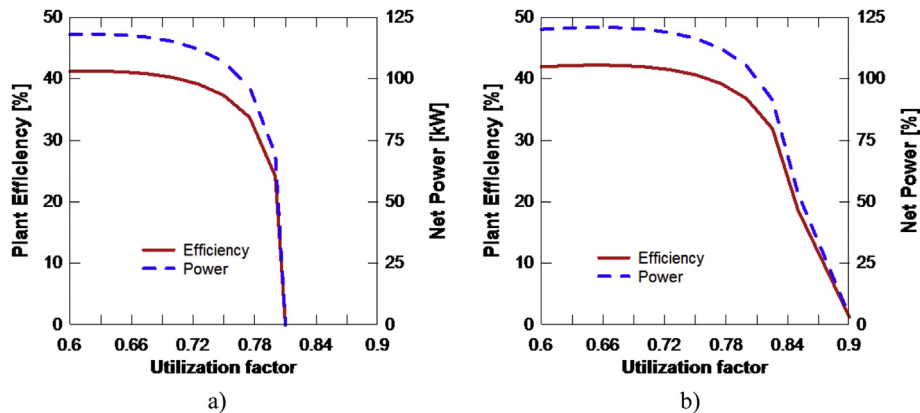


Fig. 5. Plant thermal efficiency and net power production as function of utilization factor when the woodchip mass flow is fixed: a) 100 stacks and b) 150 stacks.

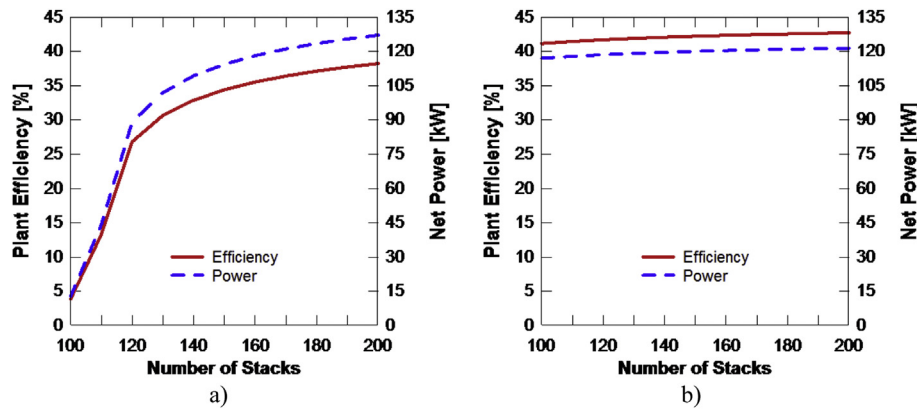


Fig. 6. Plant thermal efficiency and net power production as a function of the number of stacks when the woodchip mass flow is fixed: a) utilization factor = 0.8 and b) utilization factor = 0.65.

efficiency for the case of a utilization factor of 0.65 is higher than that for the case of 0.8. However, the power produced with a utilization factor of 0.8 could be greater than that produced with a utilization factor of 0.65 if the number of stacks is high enough (more than approximately 160). The number of stacks is directly associated with the plant investment cost, and therefore a limit must be chosen to avoid high investment costs and thereby high prices for the produced electricity. The higher the number SOFC stacks, the higher the associated investment cost would be. By closer inspection of Fig. 6, choosing 160 SOFC stacks results in higher plant efficiency than choosing 120 stacks. On the other hand, the investment cost of 160 SOFC stacks is significantly lower than that of 200 stacks. Preliminary test results showed that the selection of 160 stacks would be reasonable from the perspectives of plant efficiency and investment cost. For this reason, 160 stacks are used for the cost investigation, even though other values could be selected.

4.2. Thermoeconomical investigation

Based on the results from the thermodynamic analysis and the cost equations of each component along with the economic parameters, the operation cost per hour [\$/h] is obtained (see Table 7). The operation cost is denoted Z^{ot} , which is the component investment cost over its lifetime. The operation and maintenance costs are denoted Z^{cl} and Z^{om} , respectively.

Table 7
System and component cost rates [\$/h].

Component	Z^{cl} [\$/h]	Z^{om} [\$/h]	Z^{ot} [\$/h]
SOFC	0.8344	0.9179	1.752
Inverter	0.0071	0.0078	0.015
AP	0.0301	0.0332	0.063
Methanator	0.0356	0.0392	0.074
Syngas compressor	0.0554	0.0609	0.116
Gas cleaner	0.0572	0.0629	0.120
CP	0.5794	0.6374	1.217
Cathode compressor	0.1032	0.1136	0.217
Catalytic gas burner	0.1030	0.1134	0.216
Stirling engine	0.3965	0.4362	0.833
Water pump	0.0033	0.0036	0.007
Water heater	0.0431	0.0474	0.090
Gasifier	0.7145	0.7859	1.500
GAP	0.0629	0.0692	0.132
Steam generator	0.0573	0.0631	0.120
Steam loop blower	0.0308	0.0338	0.065
Total operating cost	2.9495	3.2444	6.194

As shown in the table, the largest cost is associated with the SOFC (1.752 \$/h), here with 160 stacks and 74 cells in each stack, followed by the gasifier (1.5 \$/h), the cathode air preheater (CP) (1.217 \$/h), which is a relatively large heat exchanger, and then the Stirling engine (0.833 \$/h). The costs are relatively low for the other components. The entire plant's operating cost is found to be 9.447 \$/h with the current inputs and assumptions.

The system's sensitivity to the fuel price is illustrated in Fig. 7, for which the system costs presented in Table 7 are used. As shown, increasing the fuel price by 0.02 \$/kWh increases the produced electricity cost by approximately 0.05 \$/kWh. For the system with the current price of woodchips described in the cost model section, a production cost of 0.1215 \$/kWh is found. The cost of the produced domestic hot water (the cost of DHW) is found to be 0.0219 \$/kWh. This low cost is due to the assumption that the cost of DHW is the cost rate of the water heater and the exergy difference of the "inflow" and "outflow" of the WH (water heater) and the Stirling engine. Hence, DHW is considered a byproduct of the plant because electricity is assumed to be the main product, and therefore the cost of the DHW will be the utilization of that heat.

There are several options for reducing the cost of the system. As suggested in Table 7, the SOFC is by far the component that contributes most to the total operating cost. Because the number of stacks is a major contributor to the cost (see Eq. (38)), reducing the number of stacks will reduce the component investment cost significantly. However, as shown in Fig. 6, reducing the number of stacks will reduce the thermal efficiency of the plant, and because

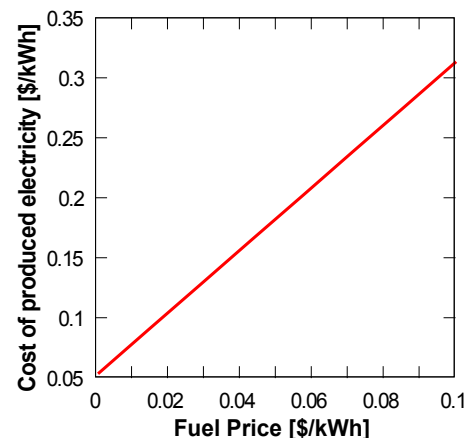


Fig. 7. Cost of produced electricity over the fuel price [\$/kWh].

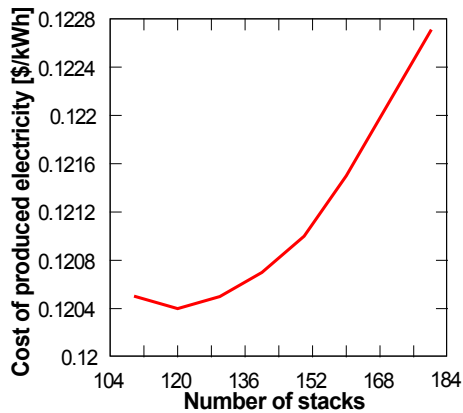


Fig. 8. Cost of produced electricity over the number of stacks in the SOFC.

the fuel cost is relatively low, the produced electricity cost will decrease with as the number of stacks decreases, see Fig. 8.

As illustrated in Fig. 8, the produced electricity cost is minimized when the number of SOFC stacks is approximately 120. This finding suggests that with the current pricing, it would be more profitable to reduce the number of stacks to approximately 120 and increase the amount of feedstock to the gasifier, which in turn decreases the plant thermal efficiency but also the cost of electricity. Thus, the number of SOFC stacks can be determined by thermoeconomical optimization. Decreasing the number of stacks to 120 results in a different plant performance, which is shown in Table 8.

The gasifier must be fed with 91.33 kg/h of wood to achieve a net power production of 120 kW. This amount of feedstock will lead to a syngas production of 180.05 kg/h, which means that the air intake into the gasifier is nearly equal to that of wood for the gasification processes. After the methanation of the syngas (to slightly increase the methane content), the power production of the SOFC plant will be approximately 98.48 kW_e. As explained earlier, the methanator increases the amount of methane in the fuel, which in turn results in endothermic internal reforming. As a result, a smaller amount of air in the cathode side will be needed to cool the SOFC to maintain the SOFC operation temperature. Thus, the power consumption of the compressor will also be reduced. The total internal power consumption will be approximately 6.10 kW. The power production from the Stirling engine is approximately 27.62 kW_e, leading to a plant thermal efficiency of 0.415 (LHV-based). The integrated gasification and SOFC topping cycle will have a thermal efficiency of 0.32 (LHV-based). Thus, applying the Stirling engine as the bottoming cycle increases the plant thermal efficiency by nearly 28%, which is significant. This

increase is due to the low thermal efficiency of the integrated gasification and SOFC plant, which leads to more energy being available for the bottoming cycle. A thermal efficiency of 41.5% may seem somewhat low, but for an integrated gasification plant producing only 120 kW_e, this efficiency is in fact sufficient. Note that an integrated gasification combined cycle has an efficiency of approximately 40% for a power output of approximately 500 MW. Note also that the SOFC fuel utilization is relatively low (0.675), as explained earlier. Furthermore, 130.52 kW_{th} of domestic hot water is produced as a byproduct.

The cost of the produced electricity is 0.1204 \$/kWh, and the cost of the produced DHW is 0.0214 \$/kWh. These costs will change if the assumptions given above change. However, the results obtained here provide a fairly good overview of the cost situation.

The obtained costs are a result of low fuel costs and high component and installation costs with respect to the plant size. As stated in energy.eu [42], the electricity price for a household with a consumption of 7500 kWh/year ($\pm 30\%$) is 0.2562 €/kWh in Denmark, and for an industry with a consumption of 2 GWh/year ($\pm 50\%$), the price is 0.0982 €/kWh. The results obtained at today's exchange rate from € to \$ lead to an electricity cost that is within the confines of what is expected when buying it from the grid; hence, the system will most likely be competitive for installation in places such as hotels and malls. Thus, the system might be cost effective when SOFCs and Stirling engines enter the commercialization phase, and their price will be close to that predicted herein. In this study, DHW is considered to be a byproduct, and the cost is obtained as 0.0214 \$/kWh, which is much lower than the price of district heating networks (0.1143 \$/kWh as provided by København Energi 2012 [43]). Thus, for in-house use, electricity is produced at a cost that is nearly competitive when fuel prices are held at the assumed level. However, for selling the electricity to the grid, the production cost is considered to be high. Compared to other renewable energy sources of similar sizes, conventional biomass gasifiers at 20–50,000 kW produce energy with costs of approximately 0.13 \$/kWh, while small-scale wind turbines produce energy at even higher rates [44].

Here, the total TCI for the installed capacity is 3432.97 \$/kW, which is high compared to other renewable energy systems.

5. Conclusions

A small-scale integrated gasification SOFC–Stirling CHP plant with a net capacity of 120 kW_e is presented and subjected to thermodynamic and thermoeconomic investigations. A rather modest plant thermal efficiency is found (0.41) after thermoeconomic optimization, partially because the parameter inputs for the different components have been chosen from a conservative viewpoint and partially because the produced syngas contains a large fraction of non-usable compounds, which in turn resulted in a lower utilization factor for the SOFC stacks.

The thermoeconomical analyses showed that by reducing the number of stacks from 160 stacks with 74 cells to 120 stacks, the cost of the produced electricity decreases at the expense of plant thermal efficiency. An electricity production price of 0.1204 \$/kWh is found with a DHW production cost of 0.0214 \$/kWh based on the assumption of component cost equations for the future pricing of SOFCs and Stirling engines when they emerge into the commercialization phase. These prices are competitive in the Danish market for in-house usage but are slightly high if the electricity is sold to the grid.

Neglecting disposal costs, the plant cost is estimated as 3433 \$/kW, which is competitive when compared to other environmentally friendly energy systems of similar sizes.

Table 8

Plant performance by thermoeconomical optimization.

Parameter	Value
Fuel consumption	91.33 kg/h
Syngas produced	180.05 kg/h
Power output, SOFC	98.48 kW
Power output, Stirling engine	27.62 kW
Total internal power consumption	6.10 kW
Thermal efficiency, SOFC cycle	0.3195
Thermal efficiency, plant	0.4149
Increase when adding bottoming cycle	29.88%
Produced heat	130.52 kW
Cost of electricity	0.1204 \$/kWh
Cost of DHW	0.0214 \$/kWh
Total TCI over installed capacity	3432.97 \$/kW

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Nomenclature

A_{hex}	heat exchanger area, m^2
$A_{i,j}$	matrix
$\dot{C}_p$	product cost rate, \$/kWh
$\dot{C}_F$	fuel cost rate, \$/kWh
c_p	specific energy cost, \$/kW
D_{bin}	binary diffusion coefficient
D_{cell}	cell diameter, m
$\dot{E}$	exergy flow rate, kW
E_{FC}	electricity from fuel cell, V
E_{Nernst}	Nernst ideal reversible voltage, V
e	specific exergy, kJ/kg
F	Faraday's constant, C/mol
f_η	efficiency correction factor
k	thermal conductivity, $\text{W}/\text{m}^2 \text{K}$
g^0	standard Gibbs free energy, J/mol
g_f	Gibbs free energy, J/mol
h	enthalpy, J/kg
h_f	enthalpy of formation, J/mol
I_{comp}	purchase cost of component k
L_{cell}	cell length, m
$\dot{m}$	mass flow rate, kg/s
$\dot{n}_{\text{H}_2}$	molar reaction rate of H_2 , mol/s
n_e	number of electron
P	power, W
p	pressure, bar
p_{H_2}	partial pressures for H_2 , bar
$p_{\text{H}_2\text{O}}$	partial pressures for H_2O , bar
Q	heat, J/s
q_i	interest rate
T	operating temperature, K
t	thickness, m
R	universal gas constant, J/K mol
RV	reversibility factor
U_F	fuel utilization factor, the actual fuel used in the SOFC cells compared to full
V	volume, m^3
V_{an}	anode porosity
W	work, W
Y	molar fraction
Z^{CI}	annual contribution to investment, \$/kWh
Z^{OM}	annual contribution to operation and maintenance, \$/kWh

Greek symbols

Δ	change/difference
ΔT_{lm}	logarithmic mean temperature difference, K
ε	effectiveness factor
η	efficiency
η_{pcy}	polytropic efficiency
v	specific volume, m^3/kg

Subscript

act	activation polarization
an	anode
aux	auxiliary
ca	cathode
conc	concentration polarization

D	destroyed
el	electricity
FC	fuel cell
L	loss
mec	mechanical
Nernst	Nernst ideal reversible voltage
ohm	ohmic polarization
ref	reference
rev	reversible
th	thermal
v	voltage

Abbreviations

AP	anode preheater
CHP	combined heat and power
CP	cathode air preheater
DHW	domestic hot water
DNA	Dynamic Network Analysis
EES	Engineering Equation Solver
GAP	gasifier air preheater
CB	catalytic burner
LHV	lower heating value
LMTD	logarithmic mean temperature difference
NTU	number of transfer units
SG	steam generator
PEC	component purchase cost
RH	room heater
SOFC	solid oxide fuel cell
TCI	total capital investment
WH	water heater

References

- [1] Sanchez D, Chacartegui R, Torres M, Sanchez T. Stirling based fuel cell hybrid systems: an alternative for molten carbonate fuel cells. *Power Sources* 2008;192:84–93.
- [2] Rokni M. Thermodynamic analysis of SOFC (solid oxide fuel cell) – Stirling hybrid plants using alternative fuels. *Energy* 2013;61:87–97.
- [3] EG&G and G Technical Services Inc. Fuel cell handbook. 7th ed. U.S. Department of Energy, Office of Fossil Energy, National Energy Technology Laboratory; 2004.
- [4] Larminie J, Dicks A. Fuel cell systems explained. 2nd ed. Chichester: John Wiley & Sons Inc.; 2004.
- [5] Komatsu Y, Kimijima S, Szmyd JS. Performance analysis for the part-load operation of a solid oxide fuel cell–micro gas turbine hybrid system. *Energy* 2010;35:982–8.
- [6] Pålsson J, Selimovic A, Sjunnesson L. Combined solid oxide fuel cell and gas turbine systems for efficient power and heat generation. *Power Sources* 2000;86(1):442–8.
- [7] Calise F, Dentice d'Accadia M, Palombo A, Vanoli L. Simulation and exergy analysis of a hybrid solid oxide fuel cell (SOFC)–gas turbine system. *Energy* 2006;31:3278–99.
- [8] Rokni M. Plant characteristics of an integrated solid oxide fuel cell and a steam cycle. *Energy* 2010;35:4691–9.
- [9] Winkler W, Lorenz H. Design studies of mobile applications with SOFC–heat engine models. *Power Sources* 2002;106:338–43.
- [10] Zhao Y, Chen J. Modeling and optimization of a typical fuel cell–heat engine hybrid system and its parametric design criteria. *Power Sources* 2009;186:96–103.
- [11] Proell T, Aichering C, Rauch R, Hofbauer H. Coupling of biomass steam gasification and SOFC–gas turbine hybrid system for highly efficient electricity generation. In: ASME turbo expo proceeding; 2004. pp. 103–12. GT2004-53900.
- [12] Bang-Møller C, Rokni M, Elmegaard B, Ahrenfeldt J, Henriksen U. Decentralized combined heat and power production by two-stage biomass gasification and solid oxide fuel cells. *Energy* 2013;58:527–37.
- [13] Bellomare F, Rokni M. Integration of a municipal solid waste gasification plant with solid oxide fuel cell and gas turbine. *Renew Energy* 2013;55:490–500.
- [14] Bang-Møller C, Rokni M. Thermodynamic performance study on biomass gasification, solid oxide fuel cell and micro gas turbine hybrid systems. *Energy Convers Manag* 2010;51:2330–9.
- [15] Rokni M. Thermodynamic investigation of an integrated gasification plant with solid oxide fuel cell and steam cycles. *Green* 2012;2:71–86.

- [16] Henriksen U, Ahrenfeldt J, Jensen TK, Gøbel B, Bentzen JD, Hindsgaul C, et al. The design, construction and operation of a 75 kW two-stage gasifier. *Energy* 2006;31(10–11):1542–53.
- [17] Ahrenfeldt J, Henriksen U, Jensen TK, Gøbel B, Wiese L, Kather A, et al. Validation of a continuous combined heat and power (CHP) operation of a two-stage biomass gasifier. *Energy Fuels* 2006;20(6):2672–80.
- [18] Hofmann PH, Schweiger A, Fryda L, Panopoulos K, Hohenwarter U, Bentzen J, et al. High temperature electrolyte supported Ni-GDC/YSZ/LSM SOFC operation on two stage Viking gasifier product gas. *Power Sources* 2007;173(1):357–66.
- [19] Elmegaard B, Houbak N. DNA – a general energy system simulation tool. In: *Proceeding of SIMS*; 2005. Trondheim, Norway.
- [20] Perstrup C. Analysis of power plant installation based on network theory [in Danish, M.Sc. thesis]. Denmark: Technical University of Denmark, Laboratory of Energetics; 1989.
- [21] Petersen TF, Houbak N, Elmegaard B. A zero-dimensional model of a 2nd generation planar SOFC with calibrated parameters. *Int J Thermodyn* 2006;9(4):161–9.
- [22] Holtappels P, DeHaart LGJ, Stimming U, Vinke IC, Mogensen M. Reaction of CO/CO₂ gas mixtures on Ni–YSZ cermet electrode. *Appl Electrochem* 1999;29:561–8.
- [23] Matsuzaki Y, Yasuda I. Electrochemical oxidation of H₂ and CO in a H₂–H₂O–CO–CO₂ system at the interface of a Ni–YSZ cermet electrode and YSZ electrolyte. *Electrochem Soc* 2000;147(5):1630–5.
- [24] Keegan KM, Khaleel M, Chick LA, Recknagle K, Simner SP, Diebler J. Analysis of a planar solid oxide fuel cell based automotive auxiliary power unit. SAE technical paper series no. 2002-01-0413.
- [25] Prentice G. *Electrochemical engineering principles*. Houston, USA: Prentice Hall International; 1991.
- [26] Achenbach E. Three-dimensional and time-dependent simulation of a planar solid oxide fuel cell stack. *Power Sources* 1994;49(1–3):333–48.
- [27] Zhu H, Kee RJ. A general mathematical model for analyzing the performance of fuel-cell membrane-electrode assemblies. *Power Sources* 2003;117:61–74.
- [28] Costamagna P, Selimovic A, Del Borghi M, Agnew G. Electrochemical model of the integrated planar solid oxide fuel cell (IP-SOFC). *Chem Eng* 2004;102(1):61–9.
- [29] Kim JW, Virkar AV. The effect of anode thickness on the performance of anode supported solid oxide fuel cell. In: *Proceedings of the sixth international symposium on SOFCs, (SOFC-VI), PV99-19*. The Electrochemical Society; 1999. pp. 830–9.
- [30] Smith JM, Van Ness HC, Abbott MM. *Introduction to chemical engineering thermodynamics*. 7th ed. Boston: McGraw-Hill; 2005.
- [31] Kromp A, Leonide A, Timmermann H, Weber A, Ivers-Tiffée E. Internal reforming kinetics in SOFC-anodes. *ECS Trans* 2010;28(11):205–15.
- [32] Winnick J. *Chemical engineering thermodynamics*. New York: John Wiley & Sons; 1997.
- [33] Reader GT. The pseudo Stirling cycle – a suitable performance criterion. In: *Proceeding of the 13th intersociety energy conversion engineering conference*, vol. 3; August 20–25, 1979. pp. 1763–70. San Diego, California.
- [34] Incropera FP, DeWitt DP, Bergman TL, Lavine AS. *Introduction to heat transfer*. 5th ed. Wiley; 2006. ISBN 978-0471457275; 2006.
- [35] Bejan A, Tsatsaronis G, Moran M. *Thermal design and optimization*. New York: John Wiley & Sons Inc.; 1996.
- [36] Arsalis A, Spakovsky M, Calise F. Thermoeconomic modeling and parameter study of hybrid solid oxide fuel cell–gas turbine–steam turbine power plants ranging from 1.5 MWe to 10 MWe. *Fuel Cell Sci Technol* 2009;6:011015.
- [37] Gary JH, Handwerk GE, Kaiser MJ. *Petroleum refining, technology and economics*. 5th ed. CRC Press; 2007. ISBN 0849370388; 2007.
- [38] Kempegowda RS, Tran KQ, Skreiberg Ø. Economic analysis of combined cycle biomass gasification fuelled SOFC systems. In: *International conference on future environment and energy (ICFEE 2011)*; 2011. pp. 134–41.
- [39] Trading Economics. Available online from: <http://www.tradingeconomics.com/denmark>; August 2012.
- [40] Ea Energy Analysis. Analysis of the market for bio energy, locally and internationally. NEI-DK-5426, sys. no.: 000492186; 2010.
- [41] Institute of Plumbing & Heating Engineering. Safe hot water temperature. Available online: <http://www.ciphe.org.uk/Global/Databyte/Safe%20Hot%20Water.pdf>; February 2014.
- [42] European Union. Available online: <http://www.energy.eu>; August 2012.
- [43] Københavns Energi. Available online: http://www.ke.dk/portal/page/portal/Erhverv/Varme/prisen/_paa/_jernvarme/_2012?page=916; August 2012.
- [44] Renewables. Global status report. REN 21, renewable energy policy for the 21st century; 2011. Paris.